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SOLIDS TRANSPORT IN ROTARY SUGAR DRYERS

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Abstract

ROTARY drum sugar drying is the last important unit operation used in the manufacture of raw sugar. In order to improve their design, operation and control, it is helpful to model their dynamic behaviour. One important aspect involves predicting the rate at which solids are conveyed through the dryer, so that hold-up of crystalline material can be better predicted. It is also important to accurately predict the amount of sugar presented to the oncoming air stream in the falling curtain, so that meaningful heat and mass transfer relations can be developed. This paper presents a dynamic model of solids transport through a rotary sugar dryer. The model is developed by assuming a tanks-in-series/parallel arrangement, akin to chemical reaction engineering systems. The use of two tanks connected in parallel allows the separation of sugar undergoing drying from the sugar particles at rest in the flights. This approach allows the prediction of air phase and rolling sugar hold-up, and residence time distribution within the dryer. Correlations to account for overloaded dryers and the impact of airflow are included. The model is based on a tracer study performed on a 100 t/h industrial dryer.

Introduction

Rotary dryers are used extensively in a number of industries for the drying of bulk particulate solids, such as raw sugar. As these dryers are most often used in the final stages of production, variations in the performance of these units can have significant impacts on the quality of the final product. Unfortunately, the mechanisms of sugar drying are complex and the impacts of operational changes on drying and cooling are not intuitive. This makes the control of flighted rotary dryers difficult, and leads to the requirement for well-validated models of drying.

In order to control the operation of a rotary dryer, it is necessary to gain an understanding of the physical phenomena occurring during operation. There are two distinct operations that occur within a dryer, namely, the transport of solids and heat and mass transfer between the solids and air. These processes are related through the dryer residence time and the cascading sugar hold-up. This paper outlines the development and validation of a dynamic model for the solids transport which occurs within a rotary dryer.

A number of approaches have been taken to modelling the solids transport phenomena occurring with rotary dryers. The foundation work in this area was performed by Friedman and Marshall (1949) and Prutton *et al.* (1942) in the form of empirical correlations between the dryer geometry and the mean residence time (τ) for the dryer. These equations are still commonly used in the design and prediction of solids transport in rotary dryers. The Friedman and Marshall equation is shown in equation 1, where M_T is the total mass holdup in the dryer; *D*, *L*, and *S* are the diameter, length and slope of the dryer respectively; *R* is the rotational speed and *F* and *A* are the solids and gas feed rates to the dryer.

K is an empirical constant characterising the effect of the airflow rate through the dryer on the holdup.

$$M_T = \frac{0.23LF}{R^{0.9}DS} + KA \qquad (1)$$

Alternatively, the Friedman and Marshall equation can be written in terms of the mean residence time of crystals in the dryer (τ) as shown in equation 2.

$$\tau = \frac{0.23L}{R^{0.9}DS} + K\frac{A}{F} \qquad (2)$$

While empirical models provide simple predictions of the mean residence time or total hold-up for a dryer, there are a number of drawbacks to these models. Firstly, it has been shown (Cao and Langrish, 1999; Renaud *et al.*, 2000) that empirical correlations generally under-predict the observed mean residence time within a full-scale dryer.

Furthermore, air-induced dispersion and back mixing occurs within these dryers and leads to a distribution of particle residence times. Figure 1 shows an experimental particle residence time distribution (RTD) for a rotary dryer (Sheehan *et al*., 2002).

Significant dispersion, as demonstrated by the breadth and tail of the distribution, can be clearly seen. Although empirical models can be used in dynamic rotary dryer simulation (Shahhosseini *et al.*, 2000) they suffer from the accuracy limitations mentioned previously and provide little insight into the physical characteristics of the cascading process (i.e. what proportion of the holdup is undergoing drying?).

A rotary dryer can be considered as essentially a two-phase system (Matchett and Baker, 1987), consisting of an cascading airborne phase undergoing drying and a passive phase of solids resting in the flights and on the floor of the dryer. This implies that only a portion of the total mass within the dryer is actually undergoing drying at any point along the dryer. Therefore, to develop a meaningful model for heat and mass transfer, a prediction of the mass in the airborne phase is required.

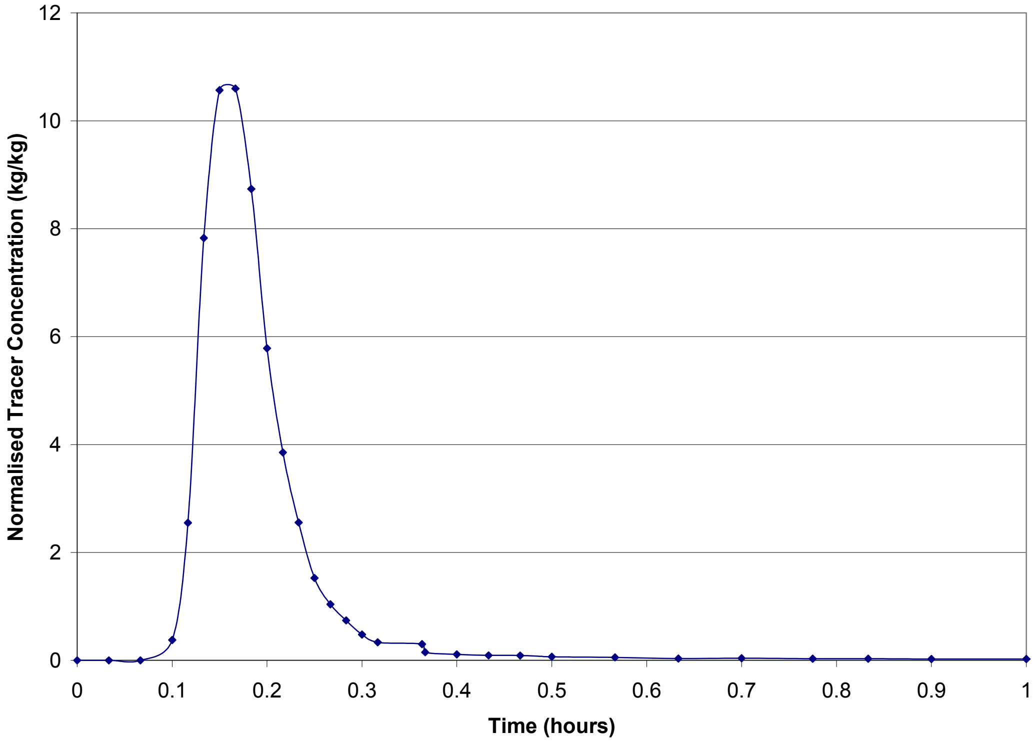

Fig. 1—Particle residence time distribution for a typical 100 t/h rotary dryer (Sheehan *et al.,* 2002).

One method of addressing these concerns is to develop a tanks-in-series model for a rotary dryer. Using the same approach that is commonly used to model complex chemical reactors, it is possible to develop a statistical model that can reproduce the observed residence time distribution (RTD) for the dryer. One such model was developed by Schneider *et al.* (2003) and is shown in Figure 2. The dryer model uses a series of pairs of ideal, well mixed tank reactors of equal volume exchanging mass with one another. The model uses four parameters to fit the experimental data; the number of pairs of tanks (N) and three solids transport coefficients (k_1, k_2, and k_3). The flow from each tank is equal to the mass in that tank multiplied by its transport coefficient.

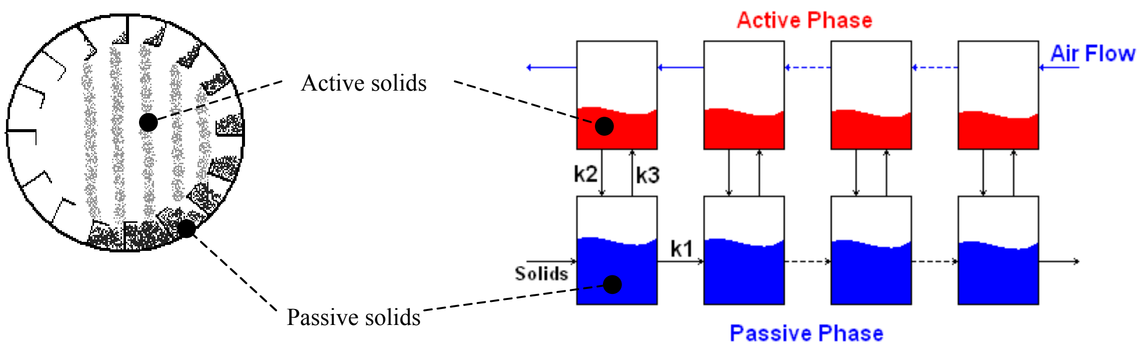

Fig. 2—Tanks-in-series/parallel model of solids transport.

The dryer model allows the separation of the two phases in the rotary dryer, and thus the prediction of the mass fraction of solids undergoing drying. In Figure 2, the *passive phase* refers to solids in the base and in the flights of the dryer. The *active phase* refers to solids cascading through the air. Due to the separation of the dryer into a number of smaller segments (N), it also allows changes in sugar properties along the dryer length to be incorporated in the model. Further details outlining the development of the dynamic mass balance equations and parameter estimation can be found in Schneider *et al*. (2003).

Model development

Before using the model, it must first be fitted to experimental data for the system in question. Schneider *et al.* (2003) used data from a single tracer experiment from an industrial sugar dryer (Sheehan *et al*., 2002), and reported that the model could accurately reproduce the observed RTD. This model was limited because it did not account for the impacts on solids transport of airflow and dryer flight capacity limitations. Two key modifications were made to account for these impacts. These were airflow rate and dryer loading.

Airflow rate correlation

Changing the airflow through the dryer will not only affect the drying of solids, but also the magnitude of the drag forces at work on the falling particles. In a counter-current dryer, such as those used to dry raw sugar, increased airflow rates will result in increased drag forces and more back mixing in the dryer. This in turn will result in a greater mean residence time. In terms of the dryer model, this effect is controlled by changing the axial transport coefficient, k_1.

At steady-state, and under design conditions[1], the axial transport coefficient is defined as $k_1 = \frac{F_d}{m_{p,d}}$ where F_d is the design solids feed rate to the dryer, and $m_{p,d}$ is the holdup of solids in a single passive tank. As the dryer is represented by a discrete number of 'slices', $m_{p,d}$ can be calculated by dividing the total passive hold-up of the dryer by the number of tanks, thus $k_1 = \frac{F_d N}{m_{p,d}}$ (Equation 3).

A steady-state mass balance on an active cell shows that $\frac{k_2}{k_3} = \frac{m_a}{m_p}$ where m_p and m_a are the solids hold-up within a passive and active cell respectively. Taking the sum across all cells it can be shown that $M_T = \left(1 + \frac{k_2}{k_3}\right) M_p$ (Equation 4) where M_p is the total passive holdup within the dryer.

Combining Equations 1, 3 and 4 gives $k_1 = \left(1 + \frac{k_2}{k_3}\right) \frac{NF_d}{M_{T,0} + KA}$ (Equation 5) where $M_{T,0}$ is the first term in Equation 1 (total passive holdup at zero airflow rate).

[1] A design-loaded dryer has enough solids in the drum to fill the flights without sugar rolling along the base of the drum (kilning).

Modelling kilning flow

When a dryer becomes design loaded (often referred to as the kilning point), the flights of the dryer are filled completely, meaning no more solids can be lifted and released into the falling curtain. Therefore, as the dryer becomes overloaded, the active phase saturates and a kilning bed of solids forms on the floor of the dryer. This creates an additional mode of transport for solids in the dryer resulting in shorter mean residence times and reduced dispersion.

A dryer operating close to its design hold-up point will swing between underloaded and overloaded conditions. To model overloaded conditions, it is necessary to define the point at which kilning occurs. A convenient definition for the kilning point is when the mass in a passive cell exceeds the total amount of solids that can be contained in the flights of that cell. Simple geometric arguments can be used to define this point.

Any additional mass will have to form a kilning bed and the rate of transport of solids between the passive and the active phase will become constant. This means that $F_{p \to a} = k_2 m_{p,d}$ where $F_{p \to a}$ is the flow rate of solids entering the active phase. As the rate of solids transport into the active phase is now constant, the mass in the active phase will rapidly reach a saturated state.

To account for the action of the kilning bed, a fourth transport coefficient was introduced, k_4, such that the flow from each passive cell equals $k_1 m_{p,d} + k_4 (m_p - m_{p,d})$. This implies that the effect of the kilning bed is linearly related to the mass of solids contained within the bed. The value of k_4 can be estimated by characterising the impact of feed rate changes on the RT or hold-up.

Results and discussion

The dryer model was implemented using a numerical modelling package (gPROMS) and the model parameters were optimised by minimising the sum of the errors squared between simulated data and data from an industrial tracer experiment (Sheehan *et al*., 2002).

It was determined that the optimal values were N(31), k_1(198.8 s^{-1}), k_2(3.01 s^{-1}) and k_3(25.4 s^{-1}). These values served as a first step for verification and validation of the model, prior to the development of a more rigorous approach to fitting the model. The optimised value for $\frac{k_2}{k_3}$ predicts an 11.9% active sugar curtain which is similar to reported values in the literature (Matchett and Baker,1987).

Using these optimized values, the airflow and kilning flow modifications proposed above were implemented. Figure 3 shows the results of simulating the steady-state hold-up in the dryer at various airflow rates using Equation 5.

As expected, the figure shows a linear dependence of hold-up on airflow, in agreement with the Friedman and Marshall equation (Equation 1). The effects of kilning flow can also be observed at the point where the slope of the curve changes.

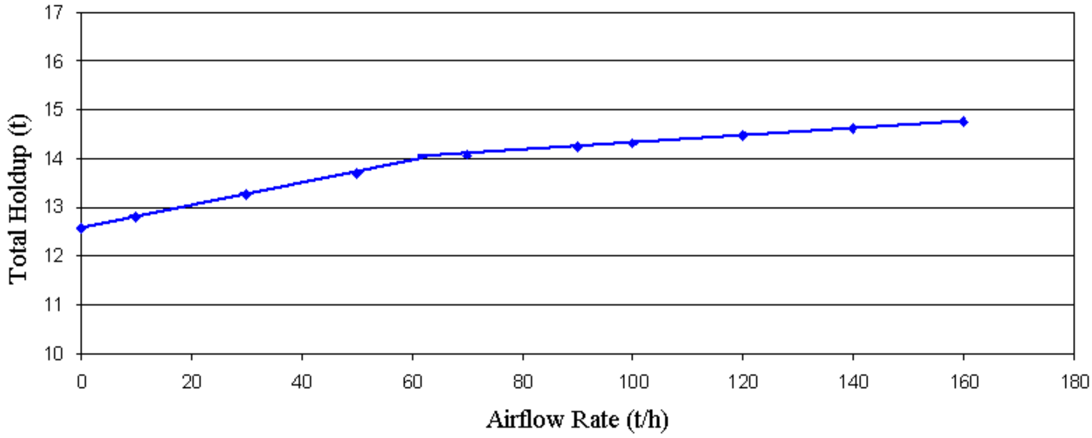

Fig. 3—Predicted dependence of total hold-up on airflow at 90 t/h feed rate.

Figure 4 shows the predicted total steady-state hold-up within the dryer against increasing solids feed rate. It can be seen that the predicted hold-up is also linearly related to the solids feed rate to the dryer. It can be seen that at approximately 92 tonnes per hour feed, the relationship between hold-up and feed rate changes, due to the kilning flow within the dryer. The effects of kilning result in a lower slope due to the increased transport of solids through the dryer.

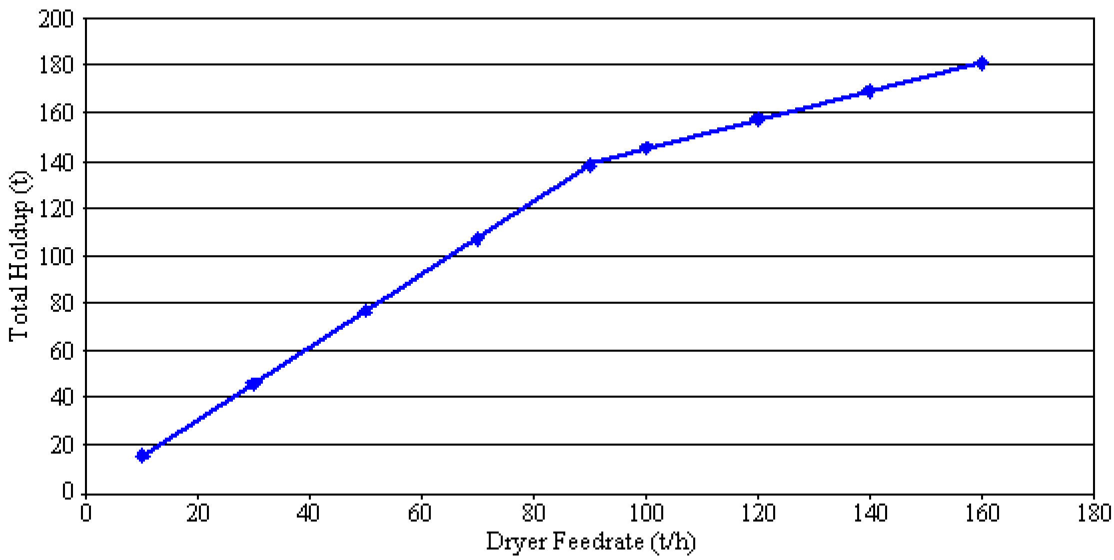

Fig. 4—Predicted relationship between total hold-up and dryer feed rate.

Conclusions

A simple dynamic model has been derived to simulate RTDs for flighted rotary dryers. The model incorporates two solids phases, aiding in the formulation of heat and mass transfer equations. The model structure is related to the dimensional and operational characteristics of a dryer and can be used to simulate different phases of operation. Minimal experimental data (RTD tracer test) are required to characterise the model. The effects of changing moisture content on sugar transport were not included but can be related through the model parameters k_2 and k_3. In order to further validate the model, tracer experiments over a wide range of operating conditions and different drum configurations are necessary and will be undertaken by the authors.

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